A Thesis

On

The Determination of Minimum Bubbling Velocity, Minimum Fluidization velocity and Fluidization Index of Fine Powders using Gas-Solid Tapered Beds.

Submitted by

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(109CH0085)

In partial fulfillment of the requirements for the degree in

Bachelor of Technology in Chemical Engineering

Under the esteemed guidance of

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National Institute of Technology Rourkela

CERTIFICATE

This is to certify that the project report entitled, "The Determination of Minimum Bubbling Velocity, Minimum Fluidization velocity and Fluidization Index of fine powders using Gas-Solid Tapered Beds." submitted by Siddhant Panda (109CH0085) in partial fulfillments for the requirements for an award of Bachelor of Technology Degree in Chemical Engineering at National Institute of Technology Rourkela is prepared by him under my supervision and guidance and this work is not been submitted elsewhere for a degree.

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ABSTRACT

Tapered beds have a velocity gradient along the axis unlike conventional beds. Tapered beds can be used in place of conventional beds to deal with problems such as slugging. Previous studies using gas-solid tapered beds have reported that stagnant bed height has no effect on minimum fluidization velocity. They have thus come up with correlations to predict minimum fluidization velocity without incorporating the initial bed height. 3 different particle systems namely Hematite, Dolomite and Limestone in the size range 20 – 80 µm were used in the study. Two tapered beds having the tapered angles of 7.36º and 9.28º were used and air was the fluidizing medium. Minimum bubbling velocities at each of the bed heights have also been observed and the fluidization index was calculated. The fluidization index is the ratio of minimum bubbling velocity to the minimum fluidization velocity. The other models for determining u_{mf} , also define u_{mf} to be that superficial velocity through the entrance at which the maximum pressure drop is obtained , i.e when the bed becomes partially fluidized from a stagnant condition whereas correlations developed here are for when the bed becomes totally fluidized (minimum full fluidization velocity) and pressure drop becomes constant. As the minimum full fluidization velocity and minimum bubbling velocity were found to increase with an increase in bed height; new correlations have been developed for predicting these values for particles in the size range of 20– 80 µm.

Keywords: Gas–solid fluidization; Minimum fluidization velocity; minimum bubbling velocity; Maximum pressure drop; Tapered fluidized bed

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NOMENCLATURE

Symbol	*Meaning*
Δp_b	pressure drop through the particle bed
$€_{mf,}$ ε_0, ε	voidage of the stagnant bed
μ	viscosity of fluidizing medium
α	Tapered Angle
u_0 , U_0	superficial velocity of the fluidizing gas
u_{mf},U_{mf}	minimum fluidization velocity
U_{mff}	minimum full fluidization velocity
Remf	Reynolds no at minimum fluidization
u_{mb}, U_{mb}	minimum bubbling velocity
ρ_g , ρ_f , ρ_s , ρ_p	gas density, fluid density, solid density, particle density
$L_{mf.}$, Hs,h_s	minimum fluidization length of bed, stagnant height of the particle bed
x_p	Mass fraction by sieve analysis
d_p	particle diameter
Ar	Archimedes no = $[g\, d_p^3\, \rho_g (\rho_s - \rho_g)]/ \mu^2$
Φ_s	sphericity of solid particle
T.A	Tapered Angle
Z	axial distance from apex of tapered vessel.

[Subscripts: 0= packed bed, 1= position of distributor, 2= top of fluidized bed)

P_{max} ,P_t	maximum pressure drop through the particle bed , total pressure drop
C_1	$= [(150(1- €_m)^2\ \mu u_0)/ (€_m^3 (\Phi_s\, d_p)^2)]$
C_2	$= [(1.75\ (1-€_{m)}\ \rho_g\ u_0^2)\ /(€_{mf}^3 \Phi_s\, \mu^2)]$
r_0 , r_1	bottom radius of the tapered bed , top radius of the tapered bed (m)
D_0 or D0, D_1 , D_c	bottom diameter of the tapered bed , top diameter of the tapered bed, Equivalent column diameter
A_t	Area of bed

CHAPTER 1

INTRODUCTION

1. INTRODUCTION

Fluidized beds have various desirable characteristics which make them of considerable use in industries. The conversion of a packed bed of solids to a fluid-like state allows rapid mixing and eases the handling operations. Isothermal conditions can be maintained in fluidized bed reactors due to this rapid mixing of solids and hence a smooth operation is possible [1].

Conventional fluidized beds face severe problems like slugging which can be tackled by introduction of baffles, operation in multistage units and imparting vibrations from time to time. The usage of tapered bed in place of a conventional cylindrical bed is an alternative technique in gas solid fluidization which can be used to tackle this problem. In tapered beds the fluid velocity varies with axial position which allows for different types of fluidizations at different positions in the tapered column [2].

Over the years various studies have been done to determine the effect the bed height has on the minimum fluidization velocity. Sau. et al [3] in their studies of gas-solid fluidization in tapered beds showed that the minimum fluidization velocity (U_{mf}) does not depend on bed height. Similar observations have been reported by Escudero and Heindel [4] who reported that for 3D bubbling fluidized beds similar to that used in their study the minimum fluidization velocity was independent of the bed height. They also reported an increase in minimum fluidization velocity with an increase in density of particles used. Khani [5] studied the hydrodynamic characteristics of tapered and mini tapered fluidized gas-solid with different cone angles for different materials. His results indicated that minimum fluidization velocity is independent of bed height. He also came up with correlations to predict minimum fluidization velocity and pressure drop across a bed. Cranfield and Geldart [6] using a 3D cylindrical bed also reported minimum fluidization velocity to be independent of bed height. Caicedo [7] reported a different observation, indicating that in gas-solid 2D beds minimum fluidization velocity was a function of height, with there being an increase in minimum fluidization velocity with an increase in bed height. This trend is attributed to the increasing friction at the wall.

Many of the important characteristics of gas-solid fluidized beds depend upon the behavior of gas bubbles which are generated near the distributor and rise through the beds growing in size [8]. Minimum bubbling velocity (U_{mb}) can be obtained by visual observation of the bed and increasing the gas velocity till the point when first bubble erupts from the free surface of the bed [9].

Fluidization index is a very important parameter, which gives us the measure of the degree to which the bed can be expanded uniformly [10]. It is the ratio of the minimum bubbling velocity to the minimum fluidization velocity. The higher the ratio the more aeratable the bed is. A high Fluidization index is also indicative of a high bed heat transfer coefficient allowing economic addition or removal of heat [11].We may obtain differences in calculated and measured values in fluidization index. This is due to particles shape and its effect on drag and minimum fluidization velocity. Thus it is better to measure U_{mf} and U_{mb} rather than to rely on correlations [10].

In this thesis the effect of bed height on minimum fluidization velocity in gas-solid tapered beds has been investigated using a 3D tapered fluidized bed. Fine powders of hematite, dolomite and limestone have been chosen. The mean diameter of hematite particles used is 20 µm. Dolomite powders having mean diameter of 58 and 76.5 µm were used for this study. Limestone of particle size 58 µm has also been used. Sau et al [3] in their studies of gas-solid fluidization in tapered beds used particles of mean diameter greater than 500 µm while predicting minimum fluidization velocity to be independent of height and hence to check whether the same applies for finer particles was a primary concern of the project. In this project the minimum velocity at which full fluidization occurs was actually observed unlike other studies where minimum velocity at which bed transitions from a stagnant bed to partially fluidized bed is observed. Two tapered bed have been used one with a tapered angle of 7.36º and other of 9.28º. In the studies particles of mean diameter in the range of 20-80 µm have been used and the effect of bed height and other parameters on minimum full fluidization velocity has been observed. Minimum bubbling velocities at each of the bed heights has also been observed and the fluidization index has been calculated.

CHAPTER 2

LITERATURE REVIEW

2. LITERATURE REVIEW

As In the course of this project we have dealt with powders therefore it is essential that we understand how powders of different particle size and density behave when fluidized. This phenomenon has been best explained by Geldart [11] in his classification of particles:-

2.1. GELDART CLASSIFICATION OF PARTICLES

Group C: Cohesive or very fine powders. Their size varies from 20 to 30 µm. In particle systems of this range, normal fluidization is extremely difficult because of the high interparticular force of attraction between the particles.

Group A: Aeratable having a small mean size and low particle density(<1.4 gm/cm^3). For this group the particle size is between 20 and 100 µm .They fluidize very easily. When the solids are fluidized the bed expands considerably (by factors of 2 and 3) before bubbles appear. The gas bubbles rise rapidly and coalesce and split frequently as they rise through the bed. When the bubbles grow to vessel diameter they turn into slugs (axial).

Group B : Sandlike, particle size 40 $u_m<d_p<500u_m$. The density is $1.4< \rho_s< 4$ g/cm^3. These solids fluidize well with vigorous bubbling. The bubbles form as soon as the gas velocity exceeds u_{mf}. Thus u_{mb}/u_{mf} = 1. Bubbles size increase roughly linearly with distance above the distributor and excess gas velocity, $(u_0\text{-}u_{mf})$.. Vigorous bubbling encourages the gross circulation of solids.

Group D: spoutable, or dense particles. The particles in this region are above 600 µm and typically have high particle densities They are difficult to fluidize. They behave abnormally giving large exploding bubbles or severe channeling.

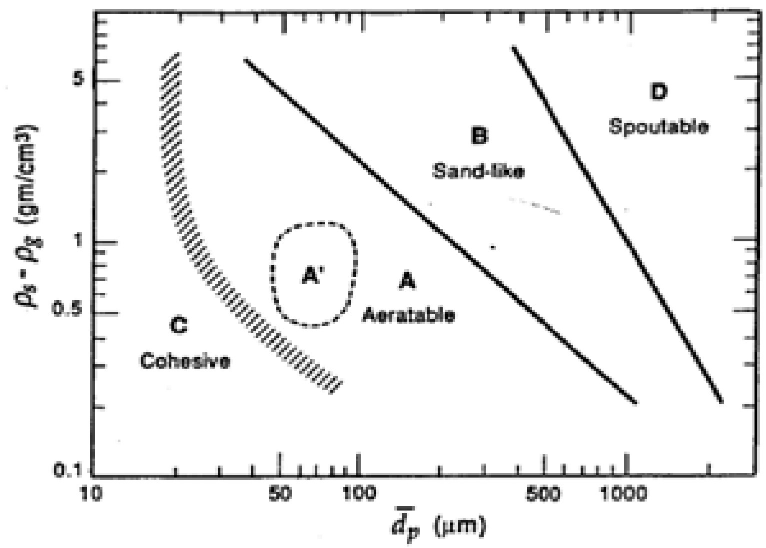

Fig 1. The Geldart classification of particles for air at ambient conditions.[Kunii et al.1992]

2.2. APEX ANGLE/TAPERED ANGLE

A relatively small apex angle is selected to accommodate the increase in gas volume with height in the case of a deep bed [12]. A large apex angle is selected to suppress slugging and to reduce bed expansion and its fluctuation effectively over a much wider range of fluidization velocities [13].

2.3. FLOW REGIMES IN A TAPERED BED

Y Peng and L.T Fan in their work on finding out the Hydrodynamic characteristics of fluidization in liquid solid tapered beds explained the various flow regimes in tapered fluidized bed in terms of pressure drop vs superficial velocity. The flow regimes were obtained during the fluidization of spherical glass beads, using water as the fluidizing medium [14].

Before starting the experimental run, the bed of glass beads were fully fluidized, subsequently the flow-rate of water was gradually reduced until the glass beads became loosely settled to form the initial fixed bed.

With an increase in flow-rate of water, the net pressure drop through the bed of particles($-\Delta p_N$) varied following the path typically described by the solid curve marked as O⇨A⇨B⇨C⇨D⇨E in the figure given in next page.

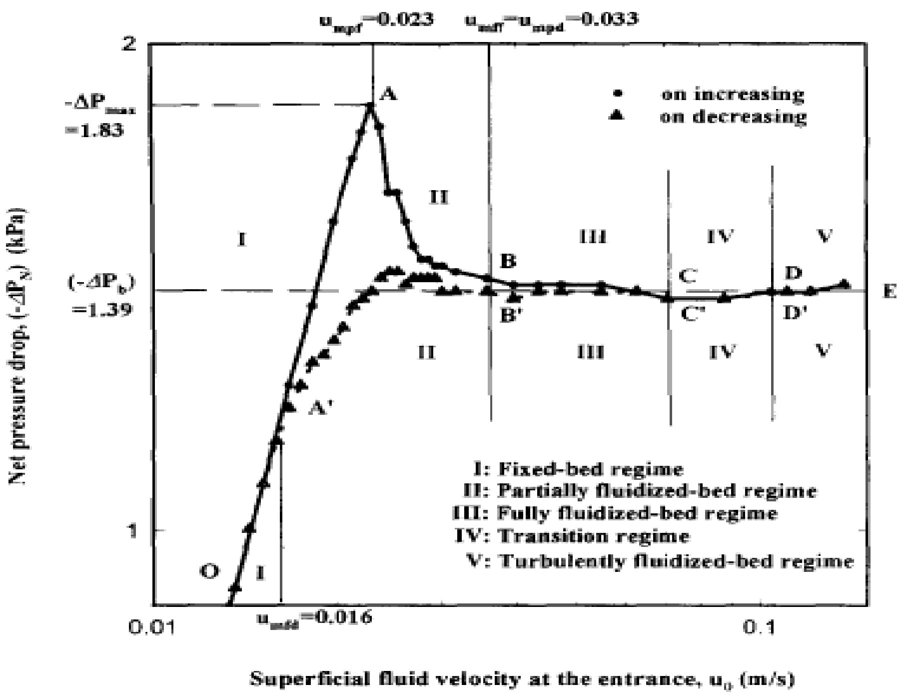

Fig 2. Pressure drop vs Superficial velocity[Peng et al.1996]

1) O⇨A, Fixed bed regime. At a low flow rate the fluid simply passes upward through the bed without disturbing the particles. The bed is maintained at a constant voidage *€*$_0$ at a height H. The magnitude of (-Δp_N*) rises* steeply with an increase in flow-rate as in the case of any fixed bed of particulates.

2) A⇨B, Partially fluidized bed regime, At point A , the particles in immediate vicinity of the distributor are lifted since flow-rate is sufficiently high , causing the formation of an empty cavity containing a relatively small no of particles next to the distributor. The cavity is unstable. A fluidized zone is formed in this cavity region. The superficial velocity at this point is known as is called the minimum velocity of partial fluidization (u_{mpf}). The bed expands as the flow-rate increases until it reaches the top of the bed. Now the Bed expands, net pressure (-Δp_N*)* drops while the flow-rate increases from max value of (-Δp_{max}*)* at point A to Δp_b at point B.

3) *B⇨C, Fully fluidized regime,* The bed reaches its critical stage of full fluidization at point B. The corresponding superficial fluid velocity through the entrance = minimum velocity of full fluidization.(u_{mff}). With only a slight increase in the flowrate beyond B, the fluidized region breaks through the top surface of the particle bed.

4) *C⇨D, Transition regime,* (-Δp_N*)* remains almost constant beyond C.

5) *D⇨E, Turbulent fluidized bed regime,* In this region particles move randomly, voidage are distributed uniformly.

2.4. PRESSURE DROP ACROSS BEDS

The pressure drop through fixed beds of length L containing isotropic solids of a single size d_p is given by Ergun's equation-

$$\frac{\Delta p_b}{L_m} = \frac{150(1-€_m)^2}{€_m^3}\frac{\mu u_0}{(\Phi_s d_p)^2} + \frac{1.75\ 1-€_m\ \rho_g u_0^2}{€_m^3\ \Phi_s d_p}$$

Pressure drop across distributors is given by :-

$\Delta p_d = (0.2\text{-}0.4)\ \Delta p_b$ [15]

The above assumption has been verified by various analyses and experiments and represents a reasonable upper bound of the required pressure drop for smooth operating conditions [1].

Maruyama et al in their paper on Fluidization in tapered vessels proposed analytical methods to predict the pressure drop in tapered fluidized vessel [17].

Pressure drop $(\Delta p_b) = (1-€_m)\ (\rho_s - \rho_g) g (Z_{2.0}^2 - Z_1^2)/(Z_2+Z_1)$

The following equation for pressure drop has been developed from Ergun's equation, which also includes the pressure drop , which is due to the K.E change in the tapered bed [3].

$$-\Delta P_{max} = \frac{C_1 H_s (D_0)\ U_0}{D_1} + \frac{[C_2 H_s D_0 (D_0^2 + D_0 D_1 + D_1^2)\ ] U_0^2}{3 D_0^2} + 1\ (U_0^2)\ \frac{[\ (D_0/D_1)^4 - 1]\ \rho_f}{2\ \varepsilon_0^2}$$

Jing et al [18] developed a model based on Ergun's equation for calculating pressure drop but neglecting the pressure drop due to the Kinetic energy change in the cylindrical bed.

The equation developed by them for nearly spherical particles is given by :-

$-\Delta P_t = C_1 U_0 H_s\ r_0/r_1 + C_2 U_0^2 H_s\ r_0/3r_1^2 (r_0^2 + r_0 r_1 + r_1^2)$

Experimentally, when using a U tube manometer the pressure drop can be determined using the formula –

$\Delta P = gR_m(\rho_a - \rho_b)$

Where,

R_m is the height difference between the two arms of the U- tube manometer,

ρ_a – Density of manometric fluid, ρ_b – Density of air.

Sau et al [3] came up with the following correlation to predict pressure drop across bed-

$$\Delta P_{max} = 7.457 * \frac{(D_1)^{0.038}}{(D_0)^{0.038}} \frac{(d_p)^{0.222}}{(D_0)^{0.222}} \frac{(H_s)^{0.642}}{(D_0)^{0.642}} \frac{(\rho_s)^{0.723}}{(\rho_f)^{0.723}}$$

Khani [5] came up with the following correlations to predict pressure drop across the bed –

$$\frac{\Delta P_{max}}{\rho_s g H_s} = [\ 106.729\ (\rho_s/\ \rho_g)^{-0.522}(d_p/\ D_0)^{0.309}(H_s/\ D_0)^{-0.379}(\cos\alpha)^{-10.86}\ ] \text{ for } 0 <= \alpha < 4.5^o$$

$$\frac{\Delta P_{max}}{\rho_s g H_s} = [\ 163.419\ (\rho_s/\ \rho_g)^{-0.524}(d_p/\ D_0)^{0.269}(H_s/\ D_0)^{-0.976}(\cos\alpha)^{-3.277}\ ] \text{ for } \alpha > 4.5^o$$

2.5. MINIMUM FLUIDIZATION VELOCITY

Minimum fluidization velocity is obtained when the pressure drop across the bed is maximum. The pressure drop increases from 0 to u_{mf}, after that it remains almost constant. The interception of the two lines is defined as the minimum fluidization velocity [7].

Kunii and Levenspiel [1] used the following analogy to predict minimum fluidization velocity:-

Drag force by upward moving gas = Weight of particles

Pressure drop across bed * cross sectional area of tube = Volume of bed* fraction consisting of solids * Specific weight of solids

$\Delta p_b A_t \quad = \quad W = \quad A_t L_{mf}(1-€_{mf})[\ \rho_s - \rho_g\ ]*g$

$\Delta p_b / L_{mf} \quad = \quad (1-€_{mf})[\ \rho_s - \rho_g\ ]*g$

In general, for isotropic solids, the minimum fluidization velocity is given by -

$$\frac{1.75(d_p u_{mf}\, \rho_g)^2}{€_{mf}^3 \Phi_s\, \mu^2} \quad + \quad \frac{150(1--€_{mf})\ (d_p u_{mf}\, \rho_g)}{€_{mf}^3 \Phi_s{}^2 \mu} \ = \ \frac{d_p^3\ \rho_g\ (\rho_s - \rho_g) g}{\mu^2}$$

Peng and Fan [14] also developed a model for determining minimum fluidization velocity

$$C_1 U_{mf} \ + \ C_2\ (\ D_0/\ D_1)^2\ U^2{}_{mf} \ - \ (1 - \varepsilon_0)(\rho_s - \rho_f) g \times \ \ [(D^2{}_0 + D_0 D_1 + D^2{}_1)\ /3D^2{}_0\ ] \ = \ 0$$

Based on the experimental data obtained by Sau et al.[3] for different types of materials , in gas-solid tapered bed and by use of dimensional analysis and estimating the constant coefficients by non linear regression they gave the following dimensionless correlation for U_{mf} :-

$$\text{Fr (Froude No)} \quad = \quad U_{\,mf}/\ (g d_p)^{0.5} \quad = \quad 0.2714(Ar)^{0.3197}(\sin \alpha)^{0.6092}(\varepsilon_0/\ \Phi_s)^{\ -0.6108}$$

They observed that maximum pressure drop and the minimum fluidization velocity increased with increase in tapered angles.

They also experimentally observed that U_{mf} was not a function of stagnant bed height in conical tapered beds. This phenomenon was also observed by Povrenovi et al [19].

Khani [5] considered the following general equation form of dimensionless correlations for calculation of minimum fluidization velocity –

$Re_{mf} = \alpha*(Ar)^b\ (dp/D_0)^c\ (\varepsilon_0/\ \Phi_s)^d(\cos\ \alpha)^e$

Non linear regression analysis was used for obtaining the constant coefficients of the correlation. From the above analysis the final equations to predict minimum fluidization velocity are –

$Re_{mf} = 7.16*(Ar)^{0.393}\ (dp/D_0)^{0.987}(\varepsilon_0/\ \Phi_s)^{-0.833}(\cos\ \alpha)^{-275.486}$, for $0<=\alpha<=4.5^o$

$Re_{mf} = 10.396*(Ar)^{0.367}\ (dp/D_0)^{0.889}\ (\varepsilon_0/\ \Phi_s)^{-0.731}(\cos\ \alpha)^{-10.347}$, for $\alpha> 4.5^o$

Sau et al and Khani in their studies took u_{mf} as the velocity at which the bed transitions from stagnant to partially fluidized and maximum pressure drop is obtained. While the correlations developed here are for the minimum velocity at which bed becomes totally fluidized (u_{mff}) [14].

Some of the studies in minimum fluidization of gas-solid 2 D tapered bed mentioned that u_{mf} decreases or increases with an increase or decrease in atmospheric pressure respectively [20].

The method of measuring u_{mf} by Caicedo et.al. was the one normally used which consists of measuring the pressure drop across the bed as a function of increasing and decreasing gas velocity as bed passes from fixed bed to fluidized bed [20]. The results of Caicedo et al. experiments showed that there is an increase in u_{mf} with an increase in bed height in 2D cylindrical beds. This trend is attributed to the increasing friction at the wall [20].

2.6 MINIMUM BUBBLING VELOCITY AND FLUIDIZATION INDEX

Abrahmsen and geldart [11] correlated the values of minimum bubbling velocity with gas and particle properties as follows-

$U_{mb}\ =\ 2.07\ \exp(0.716\ F)(x_p*\ \rho_g^{0.06}\ /\ \mu^{0.347})$
(Where F is the fraction of powder <45 µm.)

Minimum fluidization velocity for particles of size smaller than 100 μm as given by Baeyen's equation -

$U_{mf} = (\rho_p\text{-}\ \rho_g)^{0.934}g^{0.934}x_p^{1.8}/(1100*\mu^{0.87}\ *\ \rho_g^{0.066})$

Fluidization index = U_{mb}/U_{mf} = $2300\, \rho_g^{0.126}\, \mu_g^{0.523} \exp(0.176F)/[d_p^{0.8} g^{0.934} (\rho_p - \rho_g)^{0.934}]$ [10]

The higher the ratio more is the beds capacity to hold gases between minimum fluidization and bubbling point. [21]

A high fluidization index for a catalyst implies that it has a certain plasticity and can be expanded and contracted. Whereas a low fluidization index implies a brittle fluidization state in which a small change could cause a break from the uniformly fluidized catalyst to a packed bed or a bubbling regime. [21]

In some of the systems where fluidization takes place bubbles occur at velocities which is very close to u_{mf} . In some cases bubbling occurs at 3 times $u_{mf.}$. The range in which smooth trouble free fluidization occurs is extended by using fluids of high density or operating at higher pressure because the fluidization no. increases slightly with pressure and viscosity of the fluidization medium. It has been observed that powders of average particle size less than 100 μm expands uniformly without bubble formation in a limited range of gas velocity greater than minimum fluidization velocity. With materials such as fine cracking catalyst the index can be around 1.2 and the ratio varies from 1 to 2 over the range of particle sizes, with greater than 2 in some special situations [23].

We may obtain differences in calculated and measured values in fluidization index. This is due to particles shape and its effect on drag and minimum fluidization velocity. *Thus it is better to measure u_{mf} and u_{mb} rather than to rely on correlations* [23].

Singh et al. [23] predicted minimum bubbling velocity and fluidization index for gas solid fluidization in cylindrical and non cylindrical beds. Particulate fluidization exists between minimum fluidization velocity and minimum bubbling index.

Singh et al. [23] also came up with the following correlations to predict minimum bubbling velocity of different types of fluidized bed.

Cylindrical bed

$$U_{mb}=0.5231(d_p/D_c)^{1.13}(D_c/h_s)^{-0.0384}(\rho_p/\rho_f)^{0.74}$$

Semi-cylindrical bed

$$U_{mb}=0.168(d_p/D_c)^{0.994}(D_c/h_s)^{-0.1849}(\rho_p/\rho f)^{0.80}$$

Hexagonal bed

$$U_{mb}=0.15(d_p/D_c)^{0.5733}(D_c/h_s)^{-0.0887}(\rho_p/\rho_f)^{0.5384}$$

Square bed

$$U_{mb}=0.168(d_p/D_c)^{0.27}(D_c/h_s)^{-0.0132}(\rho_p/\rho_f)^{0.2825}$$

Singh et al performed experiments for dolomite, Manganese ore, Chromite ore and coal. The particle size range was ($6*10^{-4}$ m to $9*10^{-4}$ m). They observed fairly comparable value with experiments using the equations they predicted. They also observed that for identical operating conditions minimum bubbling velocity and fluidization index are maximum in case of either semi-cylindrical or hexagonal bed for most of the operating conditions and least in case of square bed. As particulate fluidization is maximum in case of semi-cylindrical beds and less in case of other beds , hence when particulate fluidization is the main requirement semi cylindrical beds are most preferred [23].

CHAPTER 3

EXPERIMENTAL SETUP, MATERIALS AND METHODS USED

3. EXPERIMENTAL SETUP, MATERIALS AND METHODS USED

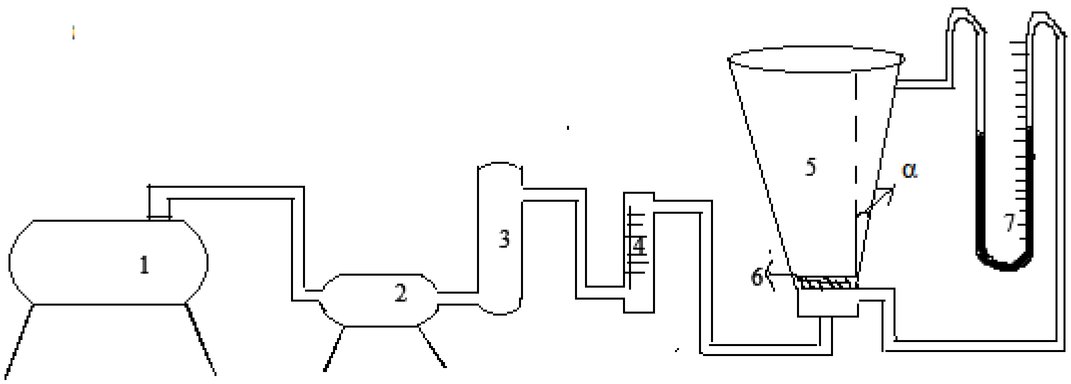

Fig 3: Experimental Setup: (1) Compressor, (2) Receiver, (3) Silica gel tower, (4) Air Rotameter,(5) Tapered bed with tapered angle α, (6) Powder bed , (7) U-tube manometer.

The first tapered bed used in this study had column bottom and top diameters of 41.5 mm and 181 mm respectively with a vertical height of 540 mm. The tapered angle was 7.36º. The second tapered bed used had column bottom and top diameters of 44.0 mm and 212 mm respectively with a vertical height of 511 cm. The tapered angle was 9.28º.

A cloth having a pore size less than 20 µm along with a perforated GI (galvanized Iron) distributor for support was used as the distributor.

A U tube Manometer having Carbon tetrachloride as the Manometric fluid (density = 1630 Kg/m^3) was used to measure the bed pressure drop. This was done by attaching separate translucent tubes on either hands of the U tube manometer; the free end of one of the tubes was attached just above the distributor. The free end of the tube attached to the other arm of the manometer was attached to the top of the tapered bed. Air was used as the fluidizing medium (At Temperature=310 K, Density = 1.17 kg/m^3, Viscosity = 1.8 $*10^{-5}$ Kg/m.s).

A silica gel tower was used to remove the moisture present in the air. A rotameter having a range of 0-50 LPM was used to measure the air flow rates.

A ball mill was used to grind dolomite chips having initial mean size in the range of 4.7-10 mm. After grinding, sieving was done using mesh sizes of BSS 100, 150, 170, 240 and 300. Two samples of powders were selected from this sieve analysis having mean diameters of 58 and 76.5 µm. Sieve analysis was also used and limestone powders having mean diameter of 58 µm was also chosen for the study.

The bulk density was measured by gently pouring the powder sample of a fixed weight in a measuring cylinder and noting its volume. The ratio of mass/poured volume gives us the bulk density. Before the start of the tests the bed was fluidized once so that the powder bed is in a loose state and is level so that the initial height can be noted properly. The particle densities were determined by the volume of water displaced method.

The voidage was measured by the formula-

Voidage (ε) = 1-(bulk density/ Particle density).

The Sphericity(Φ_s) of the powder samples were determined by using a correlation developed by Singh[22] for irregular particles which is given by –

$$\frac{(1-\varepsilon)}{\Phi_s} = -0.212 \ln d_p - 0.822$$

CHAPTER 4

RESULTS AND DISCUSSION

4. RESULTS AND DISCUSSION

The experiments were carried out for determination of minimum full fluidization velocity, minimum bubbling velocity and fluidization index of dolomite, hematite and limestone powders. Bed heights of 6, 8 and 10 cm were used and the effect of bed height on minimum bubbling and fluidization velocity and fluidization index was reported.

Before the start of each test the bed was fluidized once so that the powders were in a loosely packed state. Minimum full fluidization [14] velocities were observed as given below.

** (Below $U_{mf} = U_{mff}$ i.e. the minimum full fluidization velocity)*

4.1. FOR TAPERED ANGLE OF 7.36º

4.1.1 DOLOMITE, MEAN DIAMETER- 58 µm

For Bed Height 6 cm

Table 1

Pressure drop vs U_{mf} for bed height 6cm, dolomite 58 µm, T.A-7.36º

Pressure Drop (in Pa)	superficial velocity (in m/s)
0	0
399.06482	0.03075
399.06482	0.0615
452.2734627	0.076875
465.5756233	0.09225
465.5756233	0.107625
465.5756233	0.123
465.5756233	0.138375
452.2734627	0.1845

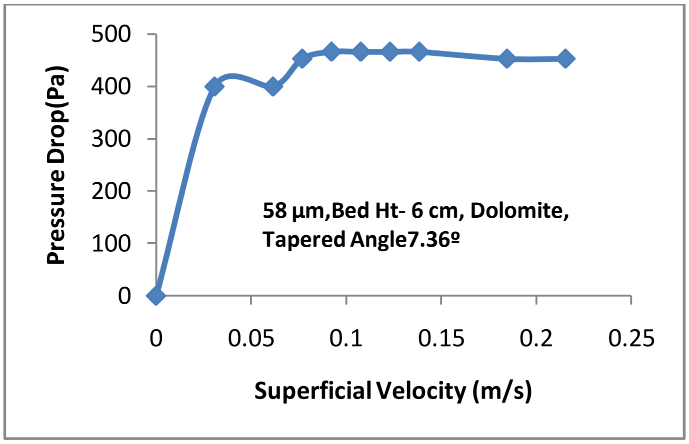

Fig 4

Pressure drop vs U_{mf} for bed height 6cm, dolomite 58 µm, T.A-7.36º

The minimum full fluidization velocity observed experimentally was 0.108 m/s and the minimum bubbling velocity observed experimentally was also 0.108 m/s.

For Bed Height 8 cm

Table 2

Pressure drop vs U_{mf} for bed height 8cm, dolomite 58 µm, T.A-7.36º

Pressure Drop (in Pa)	superficial velocity (in m/s)
0	0
532.0864267	0.03075
532.0864267	0.0615
532.0864267	0.09225
545.3885873	0.123
545.3885873	0.14145
558.690748	0.169125
545.3885873	0.199875
545.3885873	0.230625

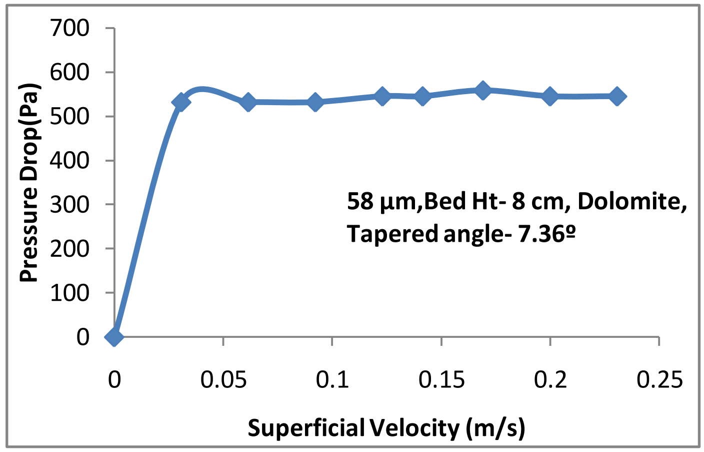

Fig 5:

Pressure drop vs U_{mf} for bed height 8cm, dolomite 58 µm, T.A-7.36º

The minimum full fluidization velocity observed experimentally was 0.123 m/s and the minimum bubbling velocity observed experimentally was also 0.123 m/s.

For Bed Height 10 cm

Table 3

Pressure drop vs U_{mf} for bed height 10cm, dolomite 58 µm, T.A-7.36º

Pressure Drop (in Pa)	superficial velocity (in m/s)
0	0
532.0864267	0.015375
665.1080333	0.046125
798.12964	0.0615
665.1080333	0.09225
678.410194	0.123
678.410194	0.138375
678.410194	0.15375
691.7123547	0.169125
718.316676	0.1845
718.316676	0.21525

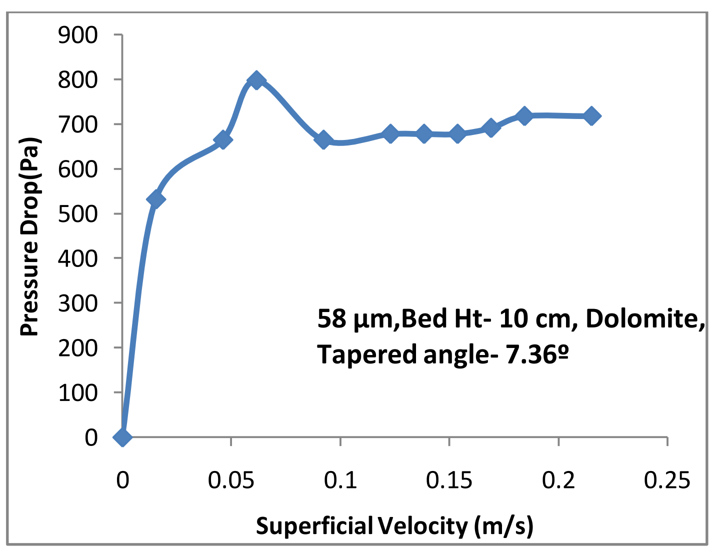

Fig 6:

Pressure drop vs U_{mf} for bed height 10cm, dolomite 58 µm, T.A-7.36º

The minimum full fluidization velocity observed experimentally was 0.169 m/s and the minimum bubbling velocity observed experimentally was also 0.169 m/s.

4.1.2 DOLOMITE, MEAN DIAMETER- 76.5 µm

For Bed Height 6 cm

Table 4

Pressure drop vs U_{mf} for bed height 6cm, dolomite 76.5 µm, T.A-7.36º

Pressure Drop (in Pa)	superficial velocity (in m/s)
0	0
226.1367313	0.003075
332.5540167	0.0046125
492.1799447	0.00615
532.0864267	0.0083025
571.9929087	0.0123
611.8993907	0.015375
505.4821053	0.046125
505.4821053	0.0615
518.784266	0.07995
505.4821053	0.0984
505.4821053	0.107625
505.4821053	0.138375
505.4821053	0.1722

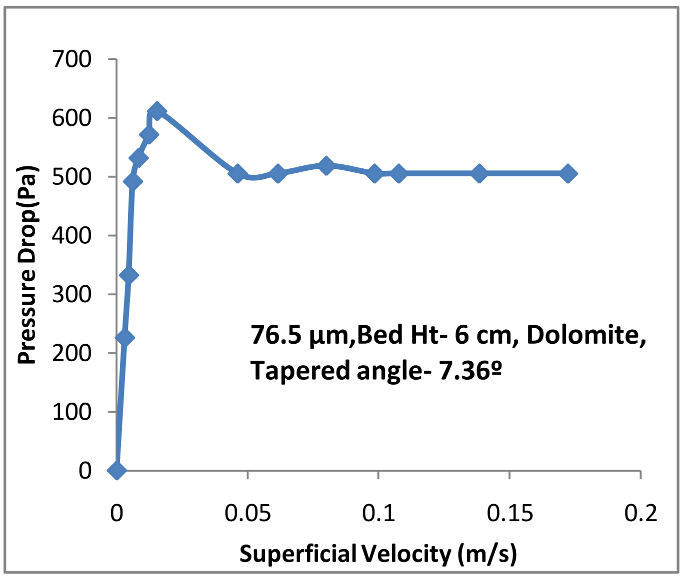

Fig 7:

Pressure drop vs U_{mf} for bed height 6cm, dolomite 76.5 µm, T.A-7.36º

The minimum full fluidization velocity observed experimentally was 0.098 m/s and the minimum bubbling velocity observed experimentally was also 0.098 m/s.

For Bed Height 8 cm

Table 5

Pressure drop vs U_{mf} for bed height 8cm, dolomite 76.5 µm, T.A-7.36º

Pressure Drop	superficial velocity
(in Pa)	(in m/s)
0	0
465.5756233	0.00615
798.12964	0.03075
638.503712	0.0615
638.503712	0.07995
638.503712	0.095325
665.1080333	0.107625
665.1080333	0.12915
665.1080333	0.15375
665.1080333	0.1845

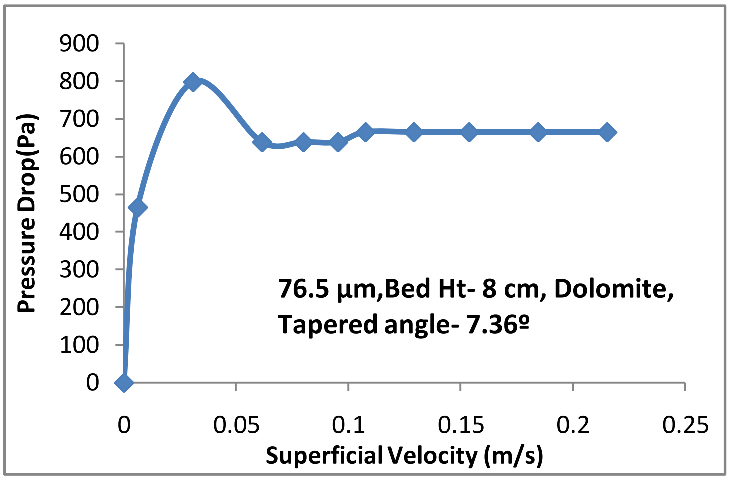

Fig 8:

Pressure drop vs U_{mf} for bed height 8cm, dolomite 76.5 µm, T.A-7.36º

The minimum full fluidization velocity observed experimentally was 0.107 m/s and the minimum bubbling velocity observed experimentally was also 0.107 m/s.

For Bed Height 10 cm

Table 6

Pressure drop vs U_{mf} for bed height 10cm, dolomite 76.5 µm, T.A-7.36º

Pressure Drop	superficial velocity
(in Pa)	(in m/s)
0	0
292.6475347	0.003075
518.784266	0.00615
731.6188367	0.015375
864.6404433	0.03075
971.0577287	0.064575
798.12964	0.09225
798.12964	0.11685
798.12964	0.138375
798.12964	0.15375

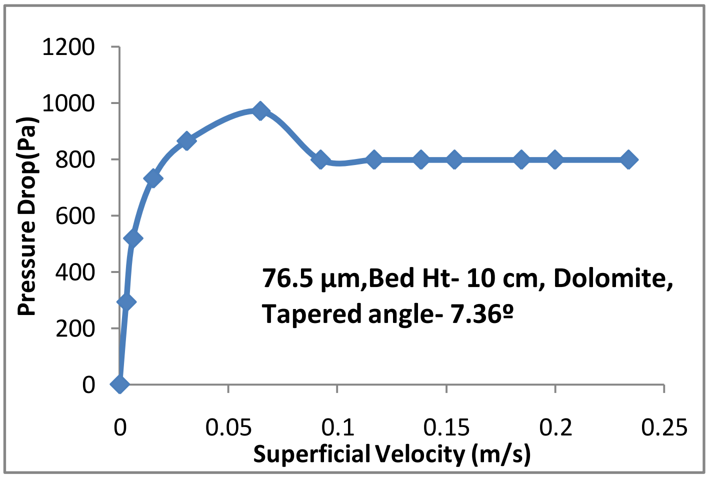

Fig 9:

Pressure drop vs U_{mf} for bed height 10cm, dolomite 76.5 µm, T.A-7.36º

The minimum full fluidization velocity observed experimentally was 0.138 m/s and the minimum bubbling velocity observed experimentally was also 0.138 m/s.

4.1.3 HEMATITE, MEAN DIAMETER- 20 μm

For Bed Height 6 cm

Table 7

Pressure drop vs U_{mf} for bed height 6 cm, Hematite 20 μm, T.A-7.36º

Pressure Drop (in Pa)	superficial velocity (in m/s)
0	0
39.906482	0.03075
79.812964	0.0615
119.719446	0.076875
133.0216067	0.09225
133.0216067	0.138375
212.8345707	0.1845
212.8345707	0.199875
212.8345707	0.21525

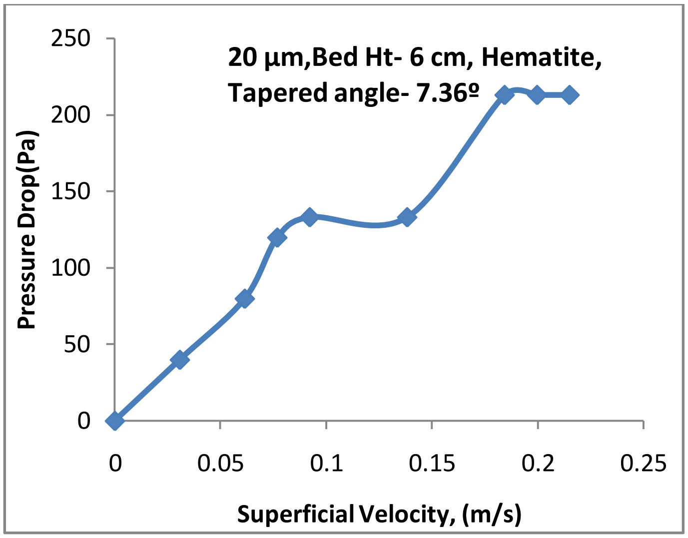

Fig 10:
Pressure drop vs U_{mf} for bed height 6 cm, Hematite 20 μm, T.A-7.36º

The minimum full fluidization velocity observed experimentally was 0.138 m/s and the minimum bubbling velocity observed experimentally was also 0.138 m/s.

For Bed Height 8 cm

Table 8

Pressure drop vs U_{mf} for bed height 8 cm, Hematite 20 µm, T.A-7.36°

Pressure Drop (in Pa)	superficial velocity (in m/s)
0	0
199.53241	0.003075
79.812964	0.00615
79.812964	0.03075
79.812964	0.0615
79.812964	0.09225
119.719446	0.123
119.719446	0.138375
119.719446	0.169125
119.719446	0.1845
119.719446	0.199875
119.719446	0.230625

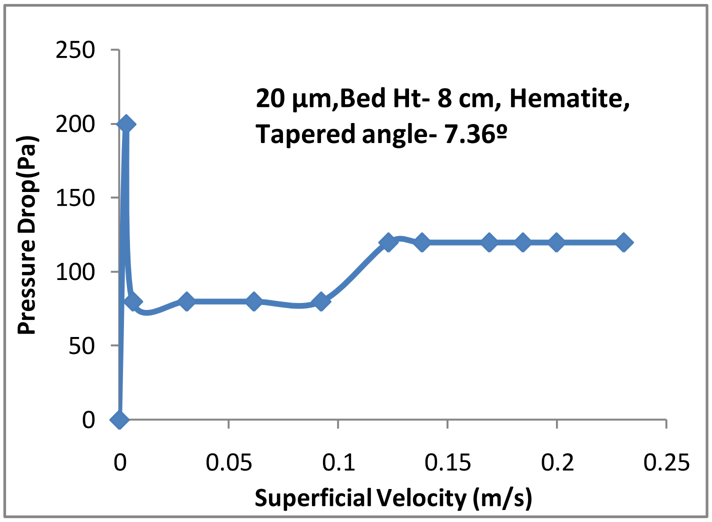

Fig 11:
Pressure drop vs U_{mf} for bed height 8 cm, Hematite 20 µm, T.A-7.36°

The minimum full fluidization velocity observed experimentally was 0.169 m/s and the minimum bubbling velocity observed experimentally was 0.169 m/s.

For Bed Height 10 cm

Table 9

Pressure drop vs U_{mf} for bed height 10cm, Hematite 20 µm, T.A-7.36°

Pressure Drop (in Pa)	superficial velocity (in m/s)
0	0
212.8345707	0.003075
172.9280887	0.03075
133.0216067	0.076875
133.0216067	0.10455
172.9280887	0.1353
212.8345707	0.169125
252.7410527	0.1845
252.7410527	0.199875
252.7410527	0.21525

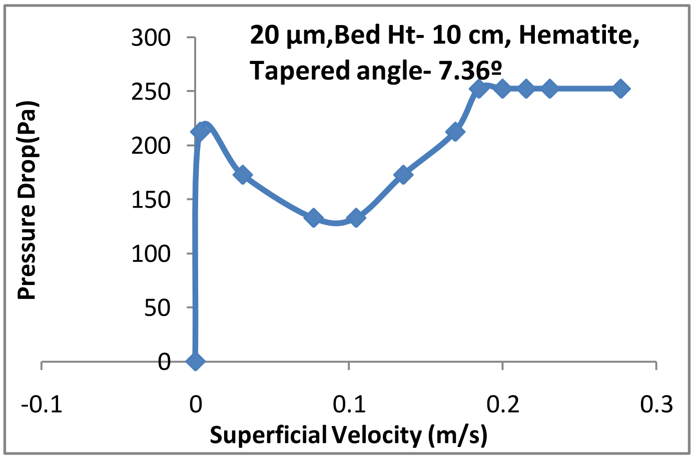

Fig 12: Pressure drop vs U_{mf} for bed height 10cm, Hematite 20 µm, T.A-7.36°

The minimum full fluidization velocity observed experimentally was 0.1845 m/s and the minimum bubbling velocity observed experimentally was also 0.1845 m/s.

4.1.4 LIMESTONE, MEAN DIAMETER- 58 µm

For Bed Height 6 cm

Table 10

Pressure drop vs U_{mf} for bed height 6cm, Limestone 58 µm, T.A-7.36°

Pressure Drop (in Pa)	superficial velocity (in m/s)
0	0
372.4604987	0.03075
385.7626593	0.0615
372.4604987	0.076875
372.4604987	0.09225
372.4604987	0.1107
372.4604987	0.15375
372.4604987	0.169125
372.4604987	0.199875

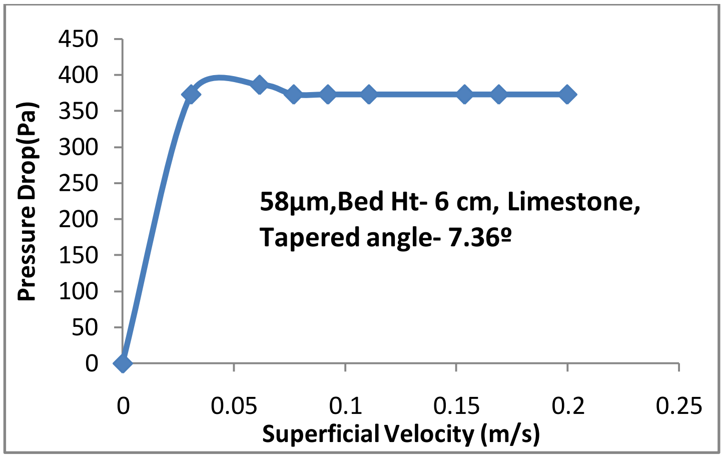

Fig 13: Pressure drop vs U_{mf} for bed height 6cm, Limestone 58 µm, T.A-7.36°

The minimum full fluidization velocity observed experimentally was 0.077 m/s and the minimum bubbling velocity observed experimentally was also 0.077 m/s.

For Bed Height 8 cm

Table 11

Pressure drop vs U_{mf} for bed height 8cm, Limestone 58 µm, T.A-7.36°

Pressure Drop (in Pa)	superficial velocity (in m/s)
0	0
518.784266	0.03075
505.4821053	0.0615
518.784266	0.09225
518.784266	0.107625
518.784266	0.123
518.784266	0.15375
478.877784	0.1845
518.784266	0.21525

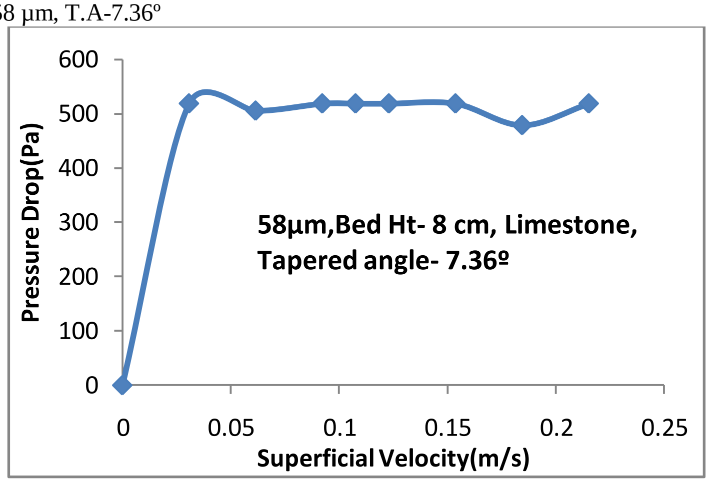

Fig 14: Pressure drop vs U_{mf} for bed height 8cm, Limestone 58 µm, T.A-7.36°

The minimum full fluidization velocity observed experimentally was 0.108 m/s and the minimum bubbling velocity observed experimentally was also 0.108 m/s.

For Bed Height 10 cm

Table 12

Pressure drop vs U_{mf} for bed height 10cm, Limestone 58 µm, T.A-7.36º

Pressure Drop (in Pa)	superficial velocity (in m/s)
0	0
638.503712	0.03075
784.8274793	0.0615
798.12964	0.083025
532.0864267	0.09225
492.1799447	0.123
505.4821053	0.138375
518.784266	0.15375
518.784266	0.1845

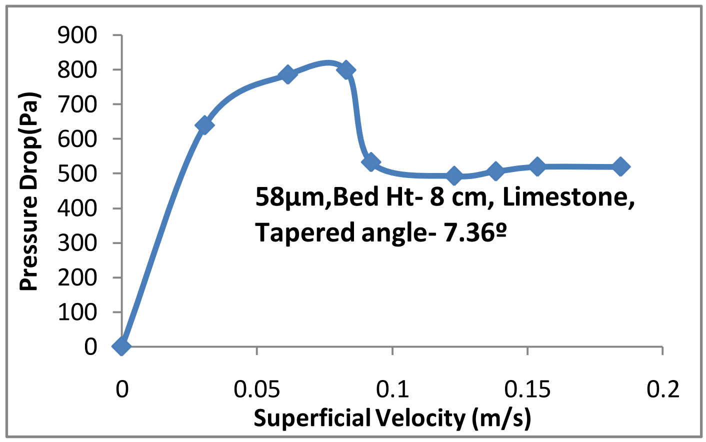

Fig 15:

Pressure drop vs U_{mf} for bed height 10cm, Limestone 58 µm, T.A-7.36º

The minimum full fluidization velocity observed experimentally was 0.123 m/s and the minimum bubbling velocity observed experimentally was also 0.123 m/s.

4.2. FOR TAPERED ANGLE OF 9.28º

4.2.1 DOLOMITE, MEAN DIAMETER- 58 µm

For Bed Height 6 cm

Table 13

Pressure drop vs U_{mf} for bed height 6 cm, Dolomite 58 µm, T.A-9.28º

Pressure Drop (in Pa)	superficial velocity (in m/s)
0	0
438.971302	0.00274
532.0864267	0.00548
372.4604987	0.0274
372.4604987	0.0685
399.06482	0.0822
425.6691413	0.1096
399.06482	0.1233
425.6691413	0.137
425.6691413	0.15344
425.6691413	0.16988

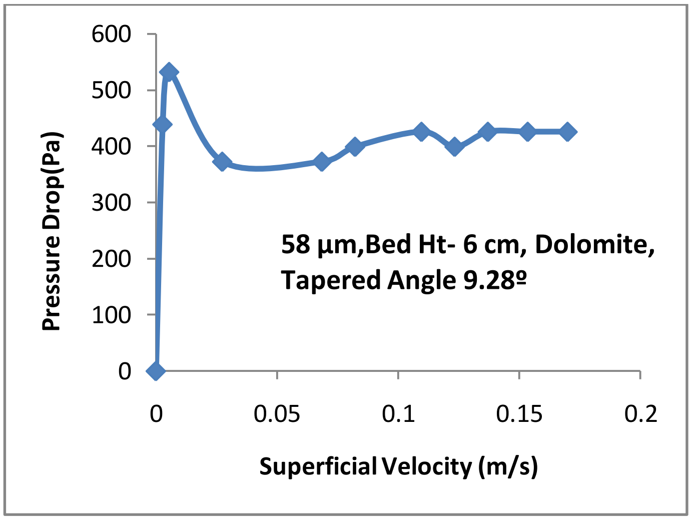

Fig 16:

Pressure drop vs U_{mf} for bed height 6 cm, Dolomite 58 µm, T.A-9.28º

The minimum full fluidization velocity observed experimentally was 0.11 m/s and the minimum bubbling velocity observed experimentally was also 0.11 m/s.

For Bed Height 8 cm

Table 14

Pressure drop vs U_{mf} for bed height 8 cm, Dolomite 58 µm, T.A-9.28º

Pressure Drop (in Pa)	superficial velocity (in m/s)
0	0
399.06482	0.00274
638.503712	0.0137
478.877784	0.0411
478.877784	0.0685
478.877784	0.09864
478.877784	0.1096
452.2734627	0.12604
452.2734627	0.137
545.3885873	0.1507
545.3885873	0.1781

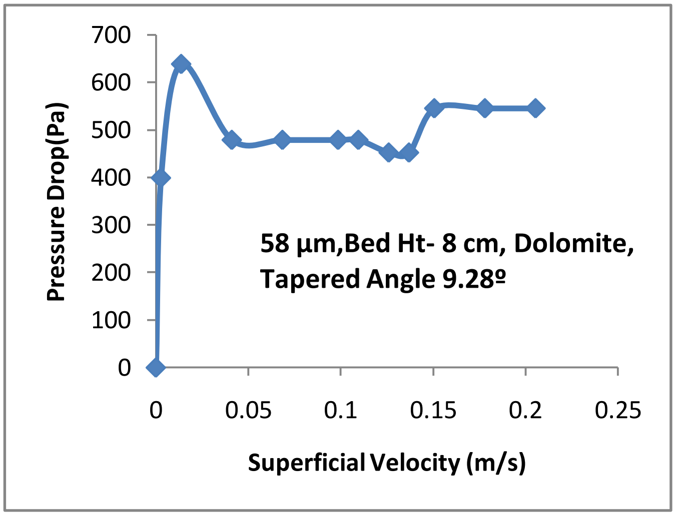

Fig 17. Pressure drop vs U_{mf} for bed height 8 cm, Dolomite 58 µm, T.A-9.28º

The minimum full fluidization velocity observed experimentally was 0.126 m/s and the minimum bubbling velocity observed experimentally was also 0.126 m/s.

For Bed Height 10 cm

Table 15

Pressure drop vs U_{mf} for bed height 10 cm, Dolomite 58 µm, T.A-9.28º

Pressure Drop (in Pa)	superficial velocity (in m/s)
0	0
199.53241	0.00274
771.5253187	0.0137
904.5469253	0.0274
984.3598893	0.0411
625.2015513	0.0548
625.2015513	0.08768
665.1080333	0.1096
585.2950693	0.1233
585.2950693	0.137
532.0864267	0.15344
532.0864267	0.17536
532.0864267	0.2055

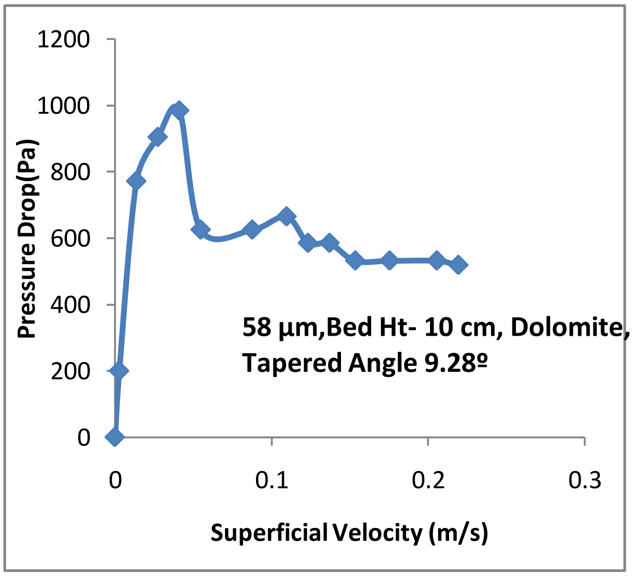

Fig. 18 Pressure drop vs U_{mf} for bed height 10 cm, Dolomite 58 µm, T.A-9.28º

The minimum full fluidization velocity observed experimentally was 0.175 m/s and the minimum bubbling velocity observed experimentally was also 0.175 m/s.

4.2.2 DOLOMITE, MEAN DIAMETER- 76.5 µm

For Bed Height 6 cm

Table 16

Pressure drop vs U_{mf} for bed height 6 cm, Dolomite 76.5 µm, T.A-9.28°

Pressure Drop (in Pa)	superficial velocity (in m/s)
0	0
372.4604987	0.00274
598.59723	0.00548
452.2734627	0.0137
465.5756233	0.0548
465.5756233	0.0685
492.1799447	0.0822
492.1799447	0.0959
492.1799447	0.1096
492.1799447	0.137
492.1799447	0.1644

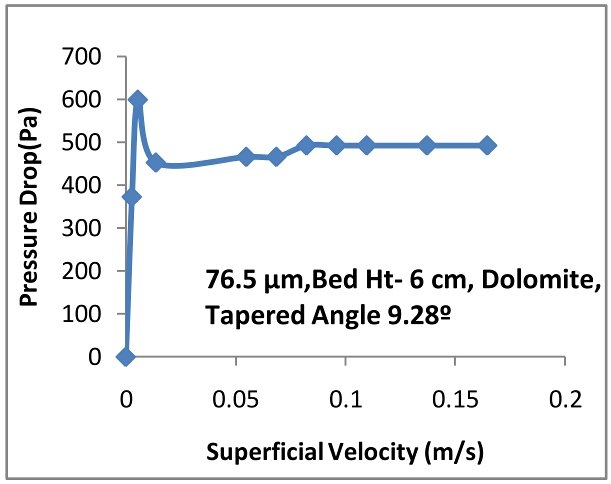

Fig 19. Pressure drop vs U_{mf} for bed height 6 cm, Dolomite 76.5 µm, T.A-9.28°

The minimum full fluidization velocity observed experimentally was 0.096 m/s and the minimum bubbling velocity observed experimentally was also 0.096 m/s.

For Bed Height 8 cm

Table 17

Pressure drop vs U_{mf} for bed height 8 cm, Dolomite 76.5 µm, T.A-9.28°

Pressure Drop (in Pa)	superficial velocity (in m/s)
0	0
345.8561773	0.00274
731.6188367	0.00548
625.2015513	0.0137
611.8993907	0.03288
611.8993907	0.07124
625.2015513	0.0959
625.2015513	0.1096
625.2015513	0.1233
611.8993907	0.1507
611.8993907	0.1781

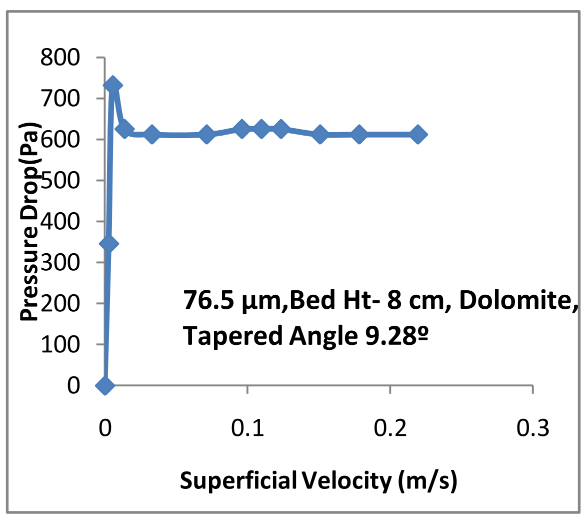

Fig. 20. Pressure drop vs U_{mf} for bed height 8 cm, Dolomite 76.5 µm, T.A-9.28°

The minimum full fluidization velocity observed experimentally was 0.096 m/s and the minimum bubbling velocity observed experimentally was also 0.096 m/s.

For Bed Height 10 cm

Table 18

Pressure drop vs U_{mf} for bed height 10 cm, Dolomite 76.5 µm, T.A-9.28°

Pressure Drop (in Pa)	superficial velocity (in m/s)
0	0
146.3237673	0.00137
492.1799447	0.00274
1050.870693	0.012056
718.316676	0.0137
718.316676	0.04384
718.316676	0.0685
718.316676	0.0822
731.6188367	0.0959
731.6188367	0.1096
731.6188367	0.1233
731.6188367	0.137

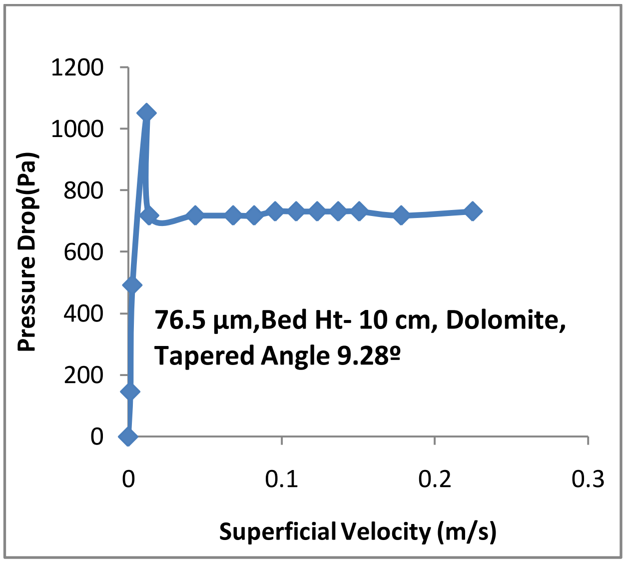

Fig.21. Pressure drop vs U_{mf} for bed height 10 cm, Dolomite 76.5 µm, T.A-9.28°

The minimum full fluidization velocity observed experimentally was 0.137 m/s and the minimum bubbling velocity observed experimentally was also 0.137 m/s.

4.2.3 HEMATITE, MEAN DIAMETER- 20 µm

For Bed Height 6 cm

Table 19

Pressure drop vs U_{mf} for bed height 6 cm, Hematite 20 µm, T.A-9.28°

Pressure Drop (in Pa)	superficial velocity (in m/s)
0	0
53.20864267	0.0137
66.51080333	0.0274
106.4172853	0.0685
119.719446	0.0822
146.3237673	0.0959
133.0216067	0.137
159.625928	0.1507
159.625928	0.1644
159.625928	0.1781
159.625928	0.1918

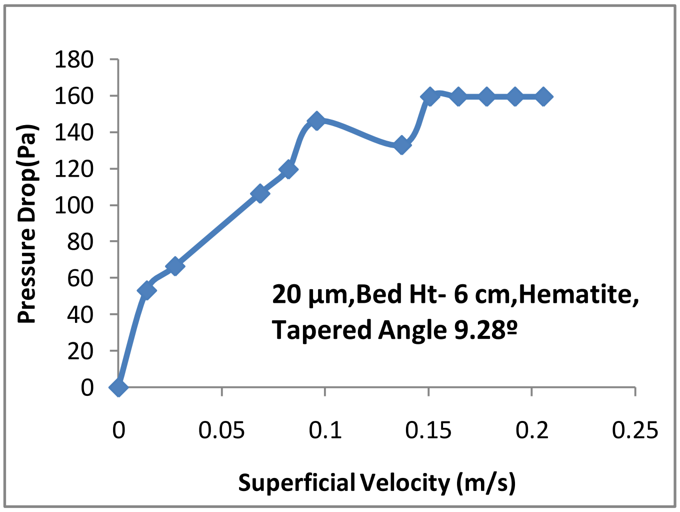

Fig 22 :Pressure drop vs U_{mf} for bed height 6 cm, Hematite 20 µm, T.A-9.28°

The minimum full fluidization velocity observed experimentally was 0.151 m/s and the minimum bubbling velocity observed experimentally was also 0.151 m/s.

For Bed Height 8 cm

Table 20

Pressure drop vs U_{mf} for bed height 8 cm, Hematite 20 µm, T.A-9.28°

Pressure Drop (in Pa)	superficial velocity (in m/s)
0	0
239.438892	0.00274
146.3237673	0.0274
146.3237673	0.0548
146.3237673	0.0822
146.3237673	0.0959
146.3237673	0.137
146.3237673	0.1644
146.3237673	0.1781
212.8345707	0.1918
212.8345707	0.20276
212.8345707	0.2192

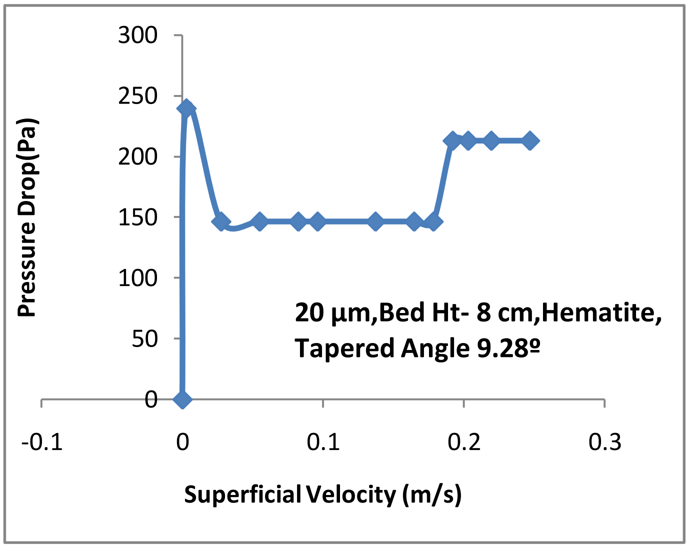

Fig.23. Pressure drop vs U_{mf} for bed height 8 cm, Hematite 20 µm,, T.A-9.28°

The minimum full fluidization velocity observed experimentally was 0.178 m/s and the minimum bubbling velocity observed experimentally was also 0.178 m/s.

For Bed Height 10 cm

Table 21

Pressure drop vs U_{mf} for bed height 10 cm, Hematite 20 µm, T.A-9.28°

Pressure Drop (in Pa)	superficial velocity (in m/s)
0	0
532.0864267	0.00274
133.0216067	0.0548
146.3237673	0.0685
199.53241	0.1096
199.53241	0.1507
212.8345707	0.1644
239.438892	0.1781
239.438892	0.1918
266.0432133	0.2055

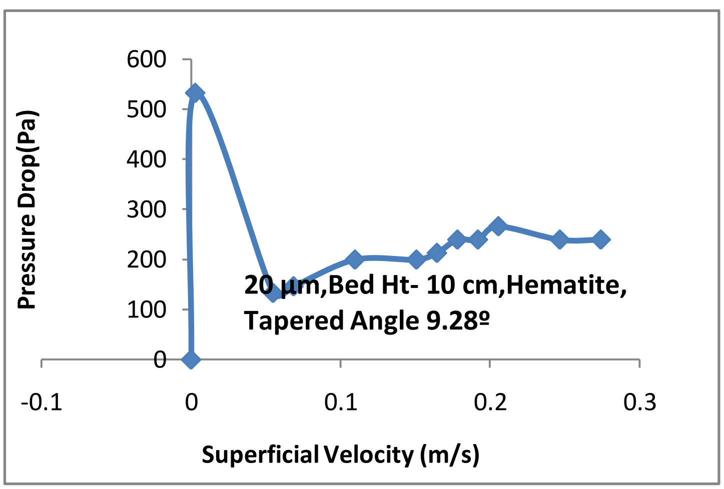

Fig 24. Pressure drop vs U_{mf} for bed height 10 cm, Hematite 20 µm, T.A-9.28°

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The minimum full fluidization velocity observed experimentally was 0.192 m/s and the minimum bubbling velocity observed experimentally was also 0.192 m/s.

4.2.4 LIMESTONE, MEAN DIAMETER- 58 µm

For Bed Height 6 cm

Table 22

Pressure drop vs U_{mf} for bed height 6 cm, Limestone 58 µm, T.A-9.28º

Pressure Drop (in Pa)	superficial velocity (in m/s)
0	0
399.06482	0.00274
266.0432133	0.0137
239.438892	0.0411
305.9496953	0.0548
345.8561773	0.0685
345.8561773	0.0822
345.8561773	0.09864
345.8561773	0.1507
345.8561773	0.1644

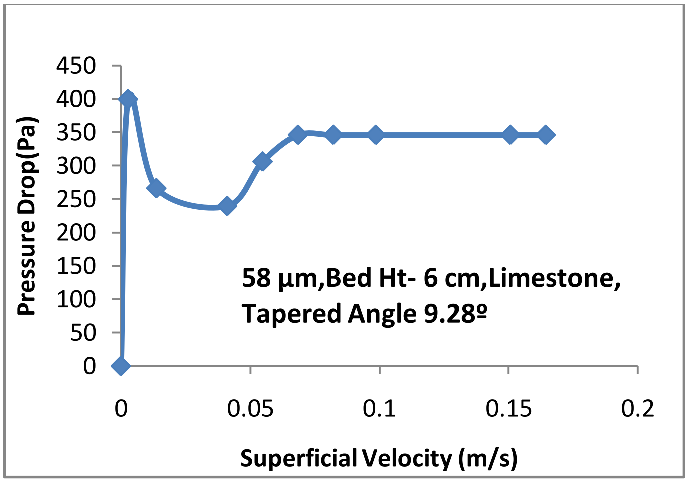

Fig 25 : Pressure drop vs U_{mf} for bed height 6 cm, Limestone 58 µm, T.A-9.28º

The minimum full fluidization velocity observed experimentally was 0.0822 m/s and the minimum bubbling velocity observed experimentally was also 0.0822 m/s.

For Bed Height 8 cm

Table 23

Pressure drop vs U_{mf} for bed height 8 cm, Limestone 58 µm, T.A-9.28º

Pressure Drop (in Pa)	superficial velocity (in m/s)
0	0
79.812964	0.00274
172.9280887	0.0137
305.9496953	0.0548
345.8561773	0.0685
372.4604987	0.0822
425.6691413	0.0959
425.6691413	0.10412
452.2734627	0.1233
452.2734627	0.137
452.2734627	0.1507

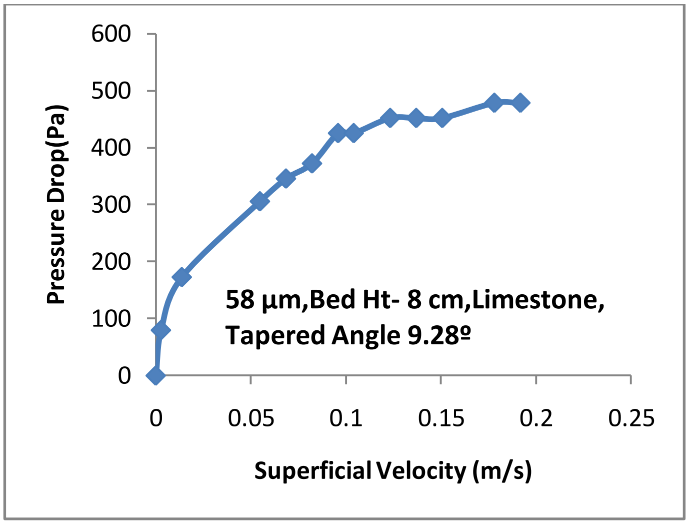

Fig 26 : Pressure drop vs U_{mf} for bed height 8 cm, Limestone 58 µm, T.A-9.28º

The minimum full fluidization velocity observed experimentally was 0.096 m/s and the minimum bubbling velocity observed experimentally was also 0.096 m/s.

For Bed Height 10 cm

Table 24

Pressure drop vs U_{mf} for bed height 10 cm, Limestone 58 µm, T.A-9.28º

Pressure Drop (in Pa)	superficial velocity (in m/s)
0	0
79.812964	0.00274
133.0216067	0.0137
438.971302	0.03836
571.9929087	0.0822
638.503712	0.10412
665.1080333	0.1233
665.1080333	0.137
665.1080333	0.1507
665.1080333	0.1644
665.1080333	0.1918
665.1080333	0.2192

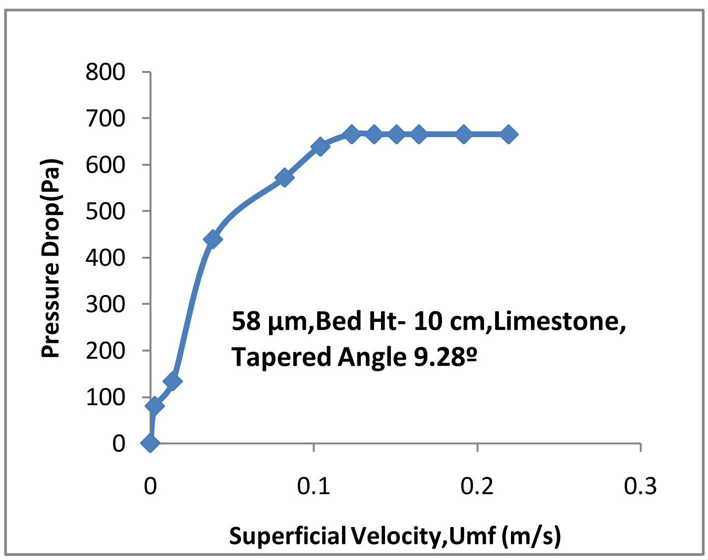

Fig 27 : Pressure drop vs U_{mf} for bed height 10 cm, Limestone 58 µm, T.A-9.28º

The minimum full fluidization velocity observed experimentally was 0.1233 m/s and the minimum bubbling velocity observed experimentally was also 0.1233 m/s.

(In all the above cases , $U_{mf} = U_{mff}$,i.e. the minimum velocity for full fluidization)*

Table 25

The particle density , bulk density , voidage and Sphericity of the particles used.

Material	Particle Diameter (µm)	Bulk Density (Kg/m^3)	Particle Density (Kg/m^3)	Voidage	Sphericity
Hematite	20	432.95	5100	0.915	0.058
Dolomite	58	899.84	2613.4	0.656	0.276
Dolomite	76.5	996.31	2613.4	0.619	0.321
Limestone	58	888.2	2812.05	0.684	0.254

Table 26

The minimum full fluidization velocity (U_{mff}), minimum bubbling velocity (U_{mb}) at different bed heights and different tapered angles of all the powders used.

			Tapered Angle – 7.36 º		Tapered Angle – 9.28 º	
Material	Particle Diameter (µm)	Bed Height (cm)	Exp.* U_{mff} (m/s)	Exp. U_{mb} (m/s)	Exp. U_{mff} (m/s)	Exp. U_{mb} (m/s)
		6	0.108	0.108	0.11	0.11
Dolomite	58	8	0.123	0.123	0.126	0.126
		10	0.169	0.169	0.175	0.175
		6	0.098	0.098	0.096	0.096

Dolomite	76.5	8	0.107	0.107	0.096	0.096
		10	0.138	0.138	0.137	0.137
Hematite	20	6	0.138	0.138	0.151	0.151
		8	0.169	0.169	0.178	0.178
		10	0.185	0.185	0.192	0.192
		6	0.077	0.077	0.0822	0.0822
Limestone	58	8	0.108	0.108	0.096	0.096
		10	0.123	0.123	0.1233	0.1233

Exp- Experimental*

From the above tabulations it is clear that the minimum full fluidization velocity and minimum bubbling velocity for each of the dolomite powders, limestone and hematite powders increases with an increase in bed height. The increase in minimum fluidization velocity and subsequently of minimum bubbling velocity may be due to increased frictional interaction with the walls [23]. The increase in minimum full fluidization velocity and subsequently of minimum bubbling velocity may also be due to electrostatic, vanderwaal's forces of attraction or capillary forces which are dominant in fine particle systems.

It can also be seen that in most of the particle systems for ex. Dolomite 58 µm, Limestone and Hematite with an increase in tapered angle there is an increase in minimum full fluidization velocity and minimum bubbling velocity; an exception is Dolomite 76.5 µm where these values remain almost the same for both the tapered beds of angle 7.36º and 9.28º.

A fluidization index value of 1 was observed in all of the cases indicating that bubbling starts from the onset of fluidization. A fluidization index value of 1 also indicates that the bed has a very less capacity to hold gases between minimum fluidization and bubbling [22].

This increase in minimum full fluidization and bubbling velocity with an increase in bed height is significant and has to be included in correlations along with the other known factors to predict them.

4.3 CORRELATION DEVELOPMENT

Sau et.al [3] and Khani [5] in their correlation development did not incorporate Hs i.e. the static bed height. They had actually determined minimum partial fluidization velocity [14] but in the correlations below minimum full fluidization velocities are being determined. In the development of correlations in this thesis hence Hs has been incorporated.

The correlation development can be proceeded by the following way:-

$Re_{mff} = f\,(Ar,\ d_p/D_0\ ,\ \varepsilon/\ \Phi_s,\ \cos\alpha,\ Hs/D_0)$

$$Re_{mff} = a * Ar^b * (\frac{dp}{D_0})^c * (\frac{\varepsilon}{\Phi s})^d * (\cos\ \alpha)^e * (\frac{Hs}{D_0})^f$$

Where a is the coefficient and b, c, d, e and f are the exponential powers.

The effect of the individual dimensionless groups on minimum full fluidization velocity has been separately evaluated by plotting log Re_{mff} (Reynolds no minimum full fluidization velocity) against log of the individual groups. Curve fitting has been done along with regression analysis and the slope obtained using the individual groups gives us the values of b, c, d, e and f. The value of a was determined by dividing the experimentally obtained Re_{mff} values against the product obtained from the RHS of the above equation and taking the weighted average.

Table 27

The values of exponential powers

b	c	d	e	f
0.286	0.670	-0.412	-1.207	0.713

The value of ‘a’ was determined to be 66.825 for hematite particle system and may be applicable to particles in the size range of around 20 µm.

The value of ‘a’ was determined to be 13.55 for the other particle systems (i.e. dolomite and limestone) and can be used for particles in the mean size range 50- 80 µm.

Thus the correlations developed are –

$$Re_{mff} = 13.55 * Ar^{0.286} * (\frac{dp}{D_0})^{0.670} * (\frac{\varepsilon}{\Phi s})^{-0.412} * (\cos\ \alpha)^{-1.207} * (\frac{Hs}{D_0})^{0.713}$$

(For particles in the size range of 50 – 80 µm)

and

$$Re_{mff} = 66.825 * Ar^{0.286} * (\frac{dp}{D_0})^{0.670} * (\frac{\varepsilon}{\Phi s})^{-0.412} * (\cos\ \alpha)^{-1.207} * (\frac{Hs}{D_0})^{0.713}$$

(For particles like hematite of size around 20 µm)

Remb (Reynolds no, when superficial velocity = U_{mb}) is also given by the same equation as U_{mb}= U_{mf} (Here U_{mf} = U_{mff} ,i.e. the minimum velocity for full fluidization[14])

Table 28

Comparison of U_{mff} values experimentally obtained and from the proposed model

Particle	d_p (μm)	α (m/s)	Bed Height(m)	U_{mff}(Exp.)	U_{mff} (m/s) by Proposed Model	Error (%)
Dolomite	58	7.36	0.06	0.108	0.092981	13.9
	58	7.36	0.08	0.123	0.114149	7.19
	58	7.36	0.10	0.169	0.133834	20.81
	76.5	7.36	0.06	0.098	0.116963	19.35
	76.5	7.36	0.08	0.107	0.14359	34.20
	76.5	7.36	0.10	0.138	0.168352	21.94
Hematite	20	7.36	0.06	0.138	0.144497	4.71
	20	7.36	0.08	0.169	0.177392	4.97
	20	7.36	0.10	0.185	0.207983	12.73
Limestone	58	9.28	0.06	0.077	0.090046	16.94
	58	9.28	0.08	0.108	0.110545	2.36
	58	9.28	0.10	0.123	0.129608	5.37
Dolomite	58	9.28	0.06	0.11	0.08617	21.66
	58	9.28	0.08	0.126	0.105787	16.04
	58	9.28	0.10	0.175	0.124046	29.12
	76.5	9.28	0.06	0.096	0.108721	13.25
	76.5	9.28	0.08	0.096	0.133471	39.03
	76.5	9.28	0.10	0.137	0.156508	14.24
Hematite	20	9.28	0.06	0.151	0.134155	11.16
	20	9.28	0.08	0.178	0.164696	7.47
	20	9.28	0.10	0.192	0.193122	0.58
Limestone	58	9.28	0.06	0.0822	0.083534	1.62
	58	9.28	0.08	0.096	0.102551	6.82
	58	9.28	0.10	0.1233	0.120251	2.47

From the above tabulations it can be seen that the proposed model performs well for the particle systems used in the study. The errors mostly lie in the range of 0-20% which is acceptable and in a few cases error higher than 20% but less than 40% is observed.

Thus the correlations developed to predict minimum full fluidization velocity and bubbling velocity are:-

$$U_{mff} = Umb = [66.825 * Ar^{0.286} * (\frac{dp}{D0})^{0.670} * (\frac{\varepsilon}{\Phi s})^{-0.412} * (\cos\ \alpha)^{-1.207} * (\frac{Hs}{D0})^{0.713} * \frac{\square}{\rho dp}]$$

(For particles like hematite of size around 20 μm)

$$U_{mff} = Umb = [13.55 * Ar^{0.286} * (\frac{dp}{D0})^{0.670} * (\frac{\varepsilon}{\Phi s})^{-0.412} * (\cos\ \alpha)^{-1.207} * (\frac{Hs}{D0})^{0.713} * \frac{\square}{\rho dp}]$$

(For particles in the size range of 50 – 80 μm)

CHAPTER 5

CONCLUSION AND FUTURE SCOPE

5. CONCLUSION AND FUTURE SCOPE

In this study the dependence of minimum full fluidization velocity and bubbling velocity on initial bed height was observed, and correlations were developed to predict U_{mff} and U_{mb} for particles in the size range of 20-80 µm. The fluidization index value of 1 was observed in all the cases.

The dependence of U_{mff} on bed height has not been reported till now, and hence this observation is new. Future studies can be planned firstly using powders of size range similar to hematite and to verify if this model holds good for these particles as well. Secondly the accuracy of this model can be checked for particles having a greater mean size than 80 µm.

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